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Assessing the potential of enhanced primary clarification to manage fats, oils and grease (FOG) at wastewater treatment works

Thomas Denis Collin[1], Rachel Cunningham[2], Mohammed Qasim Asghar[2], Raffaella Villa[3], Jitka MacAdam[1] and Bruce Jefferson[1]*

[1] Cranfield University, Cranfield, MK43 0AL, United Kingdom

[2] Thames Water Utilities Ltd., Island Road, Reading, RG2 0RP, United Kingdom

[3] De Montfort University, Leicester, LE1 9HB, United Kingdom

*Corresponding author: b.jefferson@cranfield.ac.uk

Abstract

Daily, sewage treatment works (STWs) receive large volumes of fats, oils and greases (FOG), by-products of food preparation. To increase FOG removal at STW, conventional primary sedimentation tanks (PSTs) can be enhanced using chemical coagulant or through dissolved air flotation (DAF) techniques. This work aimed to assess the potential benefits of enhanced primary treatment for FOG removal through an energy and costs analysis. To achieve this, a five-year sampling programme was conducted monthly at 15 STWs measuring FOG concentrations in crude and settled sewage (i.e. after primary treatment). In addition, two DAF pilot systems were trialled for four months and their performance, in terms of FOG removal, was assessed and compared to that of a control primary clarifier. Across the 15 STWs, influent FOG concentrations were found at 57±11 mg.L^{-1}. Chemical coagulants dosed prior to PSTs increased FOG removal rates on average to 71% whilst traditional sedimentation only achieved 50% removal. Effluent FOG concentrations were found between 12-22 mg.L^{-1} and 19-36 mg.L^{-}

^{1} respectively. By contrast, DAF achieved FOG effluent concentrations on average at 10±4 mg.L^{-1} corresponding to 74% removal from a relatively low influent concentration of 40±30 mg.L^{-1}. Thus, enhanced primary treatments have the potential to reduce organic load to secondary treatment and increase energy generation through anaerobic digestion. The overall net energy balance was estimated at 2,269 MWh.year^{-1} for the DAF compared to 3,445 MWh.year^{-1} for the chemically-enhanced PST making it a less financially attractive alternative. Yet, in the case where the works require upgrading to accommodate flow or load increases, DAF appeared as a sensible option over sedimentation offering significantly lower capital costs and footprint. In relation to FOG management, upgrading all STWs is not realistic and will require understanding where the benefits would be the highest.

Keywords: dissolved air flotation (DAF); sewage treatment works (STWs); hexane extractable material (HEM); primary sedimentation tanks (PSTs)

1 Introduction

Daily, large volumes of fats, oils and greases (FOG), by-products of food preparation, are believed to reach sewage treatment works (STWs). FOG not only causes pipe blockages within the sewerage network but also disrupts settlement and clarification processes at STWs hindering treatment efficiency (Wallace et al., 2017). In addition, FOG exerts an extra load of organic matter onto the secondary aerobic treatment stage thereby increasing the overall aeration demand. Whilst the long-chain fatty acids, which make up the majority of the FOG, can be consumed under both aerobic and anoxic conditions, kinetic studies showed that these fatty acids were degraded at a much slower rate than sugars and other substrates (Chipasa and Mędrzycka, 2006; Novak and Kraus, 1973). Consequently, FOG can accumulate within the reactors potentially enhancing the risk of foaming through stimulating the growth of filamentous microorganisms (Lefebvre et al., 1998). To avoid their detrimental impacts on downstream processes, FOG can be separated from the wastewater at the front end of STWs through a preliminary treatment step. In addition, the collected FOG is a rich energy source which can be valorised through anaerobic digestion with reported increases in methane yield of up to 138% with the addition of 23% of FOG on volatile solids (VS) basis to sewage sludge (Silvestre et al., 2011). However, separation of FOG through a preliminary stage is not always a viable option for the treatment of municipal wastewater as inclusion requires the installation of additional assets (Pastore et al., 2015).

In relatively recent years, enhanced primary treatment has been introduced through the use of coagulant dosing prior to sedimentation to increase solids and/or phosphorus removal to meet stricter effluent discharge consents. In addition, the use of dissolved air flotation (DAF) is being considered as an alternative process. DAF is commonly used in drinking water and industrial waste treatment and works by injecting air saturated pressurised water into the tank. This results in the formation of a large mass of small bubbles (40-60 μm) which combine with the solids reducing their density and causing them to float to the surface where they are removed (Edzwald, 2010). The technology is particularly effective against low density solids and hence it is posited to offer real potential for FOG removal at STW. To illustrate, in FOG-rich industrial wastewaters, removal levels of 89 and 98% have been reported from slaughterhouse effluents (Al-Mutairi et al., 2008; Karpati and Szabo, 1984; Travers and Lovett, 1985), 60% from dairy wastewaters (Monroy et al., 1995) and up to 97% in effluents from meat-manufacturing plants (El-Awady, 1999). In comparison, only a few studies have reported FOG removal efficiencies using DAF in urban wastewaters, ranging from 28% up to 72% (Kuo and Goh, 1992; Levy et al., 1972). The paucity of reported municipal cases reflects the combination of increased maintenance/operational complexity and energy demand associated with bubble generation. However, since then, technologies have become available with more optimised recycle systems and new methods of forming microbubbles (Crossley and Valade, 2006).

This work aimed to assess the potential benefit of enhanced primary treatment on FOG removal in municipal wastewater and establish the energy and operating cost basis for its potential implementation at STWs. To achieve this, an extensive sampling programme was conducted at 15 STWs over a five-year period and FOG concentrations were measured in crude and settled sewage (i.e. after primary treatment). The sites were predominantly traditional sedimentation tanks with four sites upgraded to include chemical dosing. In addition, two DAF pilot systems were trialled for four months and their performance, in terms of FOG removal, was assessed and compared to that of a control primary clarifier. Originally intended to be installed on one of the sites monitored, the DAF plants were finally trialled on a different STW due to site restraints. The results from both the extensive monitoring and the DAF plants were utilised in an economic analysis to assess the potential of advanced techniques as an alternative to conventional primary sedimentation tanks (PSTs) for FOG removal.

2 Material and methods

2.1 Process operation

Spot samples of crude sewage prior to treatment were taken monthly at 15 STWs owned by Thames Water Utilities Ltd. (TWUL) (sites 2 to 16) as part of a routine sampling. Settled sewage samples were also collected after PSTs (Table 1). Removal rates were calculated from averaged concentrations and presented with their associated propagation of uncertainties. This sampling programme was conducted over a period of five years (2013 to 2018). For each site, the surface overflow rate (SOR) was calculated as follows:

$$SOR = \frac{DWF}{A} \tag{1}$$

where: the average daily flow received at STWs is expressed as the dry weather flow (DWF) in $m^3.d^{-1}$ and A is the surface area of primary clarifiers in m^2.

Desludging from all primary treatments was achieved based on cycles controlled by timer. Ferric sulphate was dosed at concentrations around 30 $mg.L^{-1}$ (based on TWUL asset standards) upstream of the primary treatment for phosphorous removal at sites 1-5 and was posited enhancing FOG removal at these sites.

To investigate the performance of flotation techniques to remove FOG, two DAF pilot-scale systems were trialled at a municipal STW with a population equivalent (PE) of 20,090 (site 1). Unlike sites 2 to 16, STW 1 did not have any access restriction and was therefore selected for this trial. The primary treatment of wastewater was achieved via three parallel chemically-enhanced (CE) PSTs; ferric sulphate was dosed at around 25 $mg.L^{-1}$ for phosphorus removal into the PST distribution chamber. One of the PSTs was used as a control for this study. To reduce the amount of coagulant dosed into the pilot plants, a baffle was installed near the feeding point in the distribution chamber (Figure 1). As a consequence, dosing in the control-PST was reduced. Auto-desludging pumps were run for 5 minutes every 3 hours.

Crude sewage was pumped from the control-PST feeding point into a balance tank from where sewage was fed to both flotation units (Figure 1). The main differences between the two pilot-scale systems were the operation of DAF2 at a lower water pressure of 3.5 bar, and DAF1 being fitted with lamella plates increasing its effective surface area from 0.7 m^2 to 2.9 m^2 (Table 2). DAF effluents were discharged into the drain and recirculated through the treatment works.

For coagulation/flocculation, two different polyacrylamide-based polyelectrolyte aids were used. Flocculant A, recommended by one of the pilot plant manufacturers, is characterised as a cationic medium charge, high molecular weight polymer. In comparison, flocculant B was characterised as high anionic charge, very high molecular weight polymer. Dilute water solutions were made from dry polymers in 200 L round tanks. These solutions were renewed daily once used up. The dosage of polymer added was 1.5 $mg.L^{-1}$. For DAF1, dilute solutions were dosed into the sewage using a peristaltic pump. Coagulation/flocculation was achieved in a tubular contact zone prior to the flotation unit. In the case of DAF2, dilute solutions of polymer were dosed into a coagulation/flocculation tank (450 L) equipped with a mixer whose rotation speed could be adjusted.

2.2 Analytical methods

Sewage samples were analysed for total chemical oxygen demand (tCOD), biochemical oxygen demand (BOD_5) and suspended solids (SS) according to APHA methods (APHA, 2005). Total P was measured through inductively coupled plasma mass spectrometry (ICP-MS) using a Thermo Scientific^{TM} iCAP^{TM} 5200 DV. The determination of FOG in these samples was achieved by filtration, solvent extraction and gravimetry (HM Stationery Office, 1987). Briefly, wastewater samples were collected in 1 L glass bottles and filtered using a Whatman® GF/C grade filter paper. The filter paper was immersed in boiling hexane (around 50˚C), using a SOXTHERM® extraction unit, in a pre-weighted glass extraction beaker. After the solvent reduction program had run, the solvent was evaporated from beakers before being reweighed. Oil and grease concentrations were determined by weight difference. For the clarity of this paper, these results were referred to as hexane extractable material (HEM). The reporting limit of detection for this analysis was 8.2 $mg.L^{-1}$. These analyses were performed by UKAS 17025 accredited Thames Water laboratories.

During the DAF pilot-scale experiments, sewage sludge samples were regularly taken from the control-PST during auto-desludging cycles and the flotation plants. Dry solids (DS) and VS were analysed according to APHA methods (APHA, 2005). The lipids content of sewage sludge was measured using a modified Wiebul acid hydrolysis method (Sciantec Analytical, 2018).

To allow comparison between processes, HEM concentrations were normalised based on sludge produced (Q_S) as follows:

$$X_{HEM} = \frac{(HEM_{in} \times Q_i) - (HEM_{out} \times Q_{out})}{Q_S} \tag{2}$$

Where HEM_{in} and HEM_{out} are HEM concentrations measured in influent and effluent (g.m^{-3}), Q_i is the inlet flow (m^3.d^{-1}), and Q_{out} is the outlet flow (m^3.d^{-1}). The sludge production was calculated as follows:

$$Q_S = \frac{(SS_{in} - SS_{out}) \times Q_i}{\%DS} \tag{3}$$

Lipids concentrations measured in sludge were normalised based on Q_S:

$$X_{sludge} = \frac{(SS_{in} - SS_{out}) \times Q_i \times X_{lipids}}{Q_S} \tag{4}$$

Where SS_{in} and SS_{out} are SS concentrations measured in influent and effluent (g.m^{-3}), and X_{lipids} is the lipids concentrations in sludge (as %DS).

2.3 Economic evaluation

A case study was used to investigate the economic viability of retrofitting conventional clarifiers with DAF technologies at a hypothetical STW serving a PE of 500,000. Wastewater flow was assumed at 0.2 m^3.PE^{-1} per day (Henze and Comeau, 2008). Incoming BOD_5, SS and FOG loads, as well their associated removal rates from primary clarifiers, were estimated based on average values collected for sites 2 to 16. The CE-PST (low dose) scenario assumed lower chemical dose at 10 mg.L^{-1} would be needed only for FOG removal compared to the CE-PST (high dose) using around 30 mg.L^{-1} for phosphorous removal. The DAF scenario was based on removal rates obtained with DAF2-FlocB at 67%, 75% and 74% respectively for BOD_5, SS and HEM (i.e. removal rate achieved with lower HEM influent concentrations). The DAF – cost neutral scenario was developed assuming BOD_5, SS and HEM concentrations of 51 $mg.L^{-1}$, 74 $mg.L^{-1}$ and 10 $mg.L^{-1}$, obtained for DAF2-FlocB, would be achieved equating to removal of 77%, 81% and 82% based on average influent concentrations obtained from sites 2 to 16. The base year of this economic evaluation was 2018. Some cost data was collected in EUR and converted at the rate EUR:GBP of 0.80 (2008) and EUR:GBP of 0.88 (2018). Cost indices were used to adjust for the difference in capital costs over time, using the Chemical Engineering Plant Cost Index, and upon location based on European Construction Costs (2019). The relationship used for cost indices was as follows:

$$Cost\ A = Cost\ B \times \frac{Index\ A}{Index\ B} \tag{5}$$

Index values for equipment costs in 2008 and 2018 were 575.4 and 603.1. Location factors used for the UK, Denmark and Germany were respectively 100, 145.4 and 96.6 (European Construction Costs, 2019).

Capital expenditure (CapEx) for DAF was based on costs provided by the manufacturer of DAF2 at £1.76M for a plant treating 1,250 $m^3.h^{-1}$ of sewage, and £0.10M for the associated dosing plant. CapEx for PST was adapted from COWI A/S (2010) and estimated at £53 per PE. Capital costs were annualised over their lifetime (*n*) at an interest rate (*i*) of 2.8% (Ofwat, 2017). DAF and PST were assumed with lifetimes of 50 years whilst that of dosing plant was 10 years. The annualised cost of capital (ACC) was calculated as follows:

$$ACC = CapEx \times \frac{i\,(1+i)^n}{i\,(1+i)^n - 1} \tag{6}$$

Operational expenditures (OpEx) from STWs were based on (1) primary treatment (chemical costs and energy demand), (2) aeration (energy demand) and (3) sludge conditioning (chemical cost for thickening and dewatering and cake transportation cost). Energy generation from anaerobic digestion was calculated based on the sludge output from primary treatments and any additional FOG removed (Table 3).

3 Results and discussion

3.1 Occurrence of FOG in crude sewage

Across the 15 STWs monitored, influent HEM concentrations ranged from 38±37 $mg.L^{-1}$ (site 9) up to 77±50 $mg.L^{-1}$ (site 11) with a median measured at 59 $mg.L^{-1}$ (Table 1). Great variations between spot samples were observed ranging from the minimum detection limit (8.2 $mg.L^{-1}$) up to 340 $mg.L^{-1}$. Sites 9, 15 and 16, reported the lowest HEM concentrations and also displayed the lowest average BOD_5 concentrations respectively measured at 130±44 $mg.L^{-1}$, 206±64 $mg.L^{-1}$ and 170±76 $mg.L^{-1}$ (Table 1). Average values across the sites were consistent with previous reported FOG levels which vary between 10 and 100 $mg.L^{-1}$ (Dehghani et al., 2014; Gelder and Grist, 2015; Pujol and Lienard, 1989; Quéméneur and Marty, 1994; Raunkjær et al., 1994; Wiltsee, 1998).

The reported concentrations equate to per PE contribution of HEM from 9.1 $g.d^{-1}$ up to 18.0 $g.d^{-1}$ with a median measured at 11.0 $g.d^{-1}$ (Table 1). In the UK, FOG production rates at source (i.e. from households and food outlets) have been estimated around 17 $g.capita^{-1}.d^{-1}$ (Collin et al., 2020). To allow comparison, reported concentrations at STW require to be adjusted from the contribution of soaps, at 1.5 $g.capita^{-1}.day^{-1}$ (Ram et al., 2018), and lipids from faeces, at 4.1 $g.capita^{-1}.day^{-1}$ (Rose et al., 2015), as the use of a hexane extraction step within the procedure means that additional material will be included. Adjusting the measured data accordingly indicates that the actual contribution of FOG is within the range 3.5-12.4 $g.capita^{-1}.d^{-1}$ and a median value of 5.4 $g.capita^{-1}.d^{-1}$. It is posited that the difference reflects accumulation within the sewer network, potentially accounting for 69% of the FOG that enters the network. A stronger relationship was observed between influent HEM and BOD_5 concentrations compared to tCOD and SS concentrations. To illustrate, correlations of 0.79, 0.87 and 0.65 were determined for tCOD, BOD_5 and SS respectively. The close relationship existing between the BOD_5 and HEM could indicate a good degradation by aerobic biological organisms over a specific period of time.

3.2 Treatment performance

The lowest HEM concentrations in effluents from primary treatment were measured for site 13 at 19±12 $mg.L^{-1}$ whilst the highest were reported for site 8 at 36±24 $mg.L^{-1}$ (Table 1). Corresponding removal efficiencies across the traditional sedimentation processes ranged between 33±4 and 64±8% (median 54%). In contrast, enhancing primary treatment through chemical dosing increased FOG removal to between 64±9% up to 76±9% with a median of 73%. This equated to effluent FOG concentrations of between 12±10 and 22±14 $mg.L^{-1}$. A one-way ANOVA resulted in a F_{value} of 15.5 and a F_{crit} of 4.7 at a confidence interval of 95% (p value 0.002) such that there was a significant difference between HEM removal rates from conventional and CE-PSTs. The current results reflect a higher overall removal than observed during previous studies that reported FOG removal rates from conventional PSTs at US STWs which were between 45% (Loehr and de Navarra Jr., 1969; Murcott, 1992) and 47% (Gehm, 1942). With respect to chemical dosing, previous trials reported removal of 59% (Kuo and Goh, 1992) and 71% (Murcott, 1992). SOR were substantially higher in trials conducted by Kuo and Goh (1992) at between 1.5 $m.h^{-1}$ to 3.0 $m.h^{-1}$. However, FOG removal did not correlate with SOR (Figure 2), a finding supported by Loehr and de Navarra Jr. (1969). It is posited that removal reflects more the efficacy of dosing in relation to dosed amount and mixing conditions, observations that are commonly reported with respect to coagulation of drinking water (Fearing et al., 2004) and tertiary treatment of sewage (Murujew et al., 2020).

Similarly, tCOD, BOD_5 and SS removal efficiencies were found to be significantly higher with CE-PSTs. To illustrate, removal efficiencies across conventional PSTs ranged between 38-52% for tCOD (median 46%), 26-53% for BOD_5 (median 42%), and 49-72% for SS (median 62%). By contrast, CE-PSTs achieved tCOD removals between 56-68% (median 62%), 48-70% for BOD_5 (median 63%), and 70-81% for SS (median 72%). A stronger relationship was observed between HEM and both tCOD and BOD_5 removal rates, with correlation of 0.79 and 0.77, than with SS determined at 0.55 (Figure 3).

3.3 DAF pilot-scale experiments

Two DAF pilot-scale systems were trialled with the aim to compare their performance to that of PSTs gathered during the extensive sampling. Comparison of the control CE-PST and the three DAF trials revealed HEM removal efficiencies of 65±10%, 51±12%, 61±11% and 74±10% for the CE-PST, DAF1-FlocA, DAF2-FlocA and DAF2-FlocB (Table 4). The corresponding effluent concentrations were 14±7, 20±12, 16±8 and 10±4 $mg.L^{-1}$ from a relatively low influent concentration of 40±30 $mg.L^{-1}$. There was a significant difference between effluents from the control-PST and DAF2-FlocB at a confidence interval of 90% (ANOVA, *p*-value 0.07). Accordingly, the nature of the polymer appeared to have a significant impact on the efficacy of the processes with the best results observed for the anionic, very large molecular weight polymer. The importance of appropriate polymer selection has been reported before with charge, size and structure all known to influence the outcome as polymers are able to work through a number of different mechanisms such as charge neutralisation, steric hindrance and bridging (Murujew et al., 2020).

The levels reported for DAF2-Floc B were comparable to previous reported FOG removal rates at 72% for DAF treating municipal sewage (Kuo and Goh, 1992) or FOG-rich industrial wastewaters (Jensen et al., 2014; Monroy et al., 1995). Whilst DAF2-FlocB achieved relatively comparable performance in removing FOG as CE-PSTs, the process was operated at much higher SOR providing significant opportunities in terms of footprint reduction (Figure 2). In addition, it should be noted that chemical dosing is included to improve solids or phosphorus removal and not specifically FOG. Comparison during the trial revealed improved solids removal with the DAF compared to the CE-PST at 75±7% and 69±5% respectively but slightly poorer phosphorus removal at 49±4% compared to 54±3% respectively. Removal efficiencies of tCOD and BOD were also slightly higher for the DAF plant but the greatest difference was observed with regards to HEM.

The sludge produced from the different primary treatments had a DS level of 3.1±1.0%, 6.6±1.4%, 7.1±1.1% and 4.9±1.4% for the control-PST, DAF1-FlocA, DAF2-FlocA and DAF2-FlocB respectively. Accordingly, flocculant A appeared to be more appropriate for dewatering rather than primary removal. Lipid analysis revealed that not only was the sludge from control-PST less concentrated but it also contained fewer lipids. To illustrate, lipids concentrations as a fraction of the DS were 7.0±3.0% for the control-PST compared to 9.1±2.9% for DAF1-FlocA, 12.2±4.3% for DAF2-FlocA and 13.0±6.6% for DAF2-FlocB (Figure 4). A one-way ANOVA showed there were significant differences, at a confidence interval of 95%, in the lipids content of control-PST sludge and DAF2 flotation sludge. Comparison to literature revealed relatively low levels in the current study with reported ranges of 6.2 up to 19.4% DS with an average at 10.8% DS for primary sludge from sedimentation (Barber, 2014; Giacalone, 2017; Gonzalez, 2006) and 20.0-44.1% of DS for flotation sludge (Donoso-Bravo and Fdz-Polanco, 2013; Perez et al., 2012; Silvestre et al., 2011). Whilst a few authors have reported very high levels of up to 94.5% with a median of 31.7% in terms of DS for FOG harvested at STWs (Collin et al., 2020; Martín-González et al., 2011; Williams et al., 2012), it is posited that the low levels reported here reflect the low influent concentrations in the sewage. To verify this hypothesis, HEM removed and lipids in sludge were normalised based on m^3 of sludge produced. The DAF pilot-scale systems were found better at removing

FOG, generating between 5.7±0.8 to 8.8±1.1 kg lipids.m^{-3} sludge produced, compared to the control-PST calculated at 2.3±0.2 kg lipids.m^{-3} sludge produced (Table 4). For the DAF plants, normalised quantities of lipids in sludge represented between 82 and 101% of the quantities of lipids found in sludge confirming that the final concentrations in sludge were fed limited. Higher influent lipids concentrations would have produced a greasier sludge. In the case of DAF1, sampling before and after the screens indicated removal rates of 16%, 32% and 32% of the incoming BOD_5, tCOD and SS loads respectively. Consequently, this also had a direct impact on the sludge quality possibly reducing lipids content.

3.4 Economic evaluation

Excluding the energy demand from the process, the impact of using enhanced primary treatment in terms of the energy gain revealed a net positive change in energy of 3,460 MWh.year^{-1} and 4,801 MWh.year^{-1} for the CE-PST and DAF systems respectively compared to conventional PST (Table 5). In both cases, the majority of the benefit was observed with respect to reduction in energy demand for aeration as opposed to energy generation in anaerobic digestion. For instance, the reduction in energy demand generated by the enhanced removal of the DAF plant accounted for 70% of the total benefits. Energy generation from flotation sludge through anaerobic digestion was estimated at 7,592 MWh.year^{-1}, with FOG providing an additional 651 MWh.year^{-1}, corresponding to an increase of 19% to the baseline scenario (i.e. 6,137 MWh.year^{-1} generated with conventional primary treatment). Furthermore, the improved management of FOG contributed to 42% of the total benefits for the DAF and 52% for the CE-PST. The significantly higher energy benefit of the DAF plant is reduced by the increased energy demand for operation compared to the CE-PST at 2,555 and 38 MWh.year^{-1} respectively. The overall net energy balance is therefore 2,269 MWh.year^{-1} for the DAF compared to 3,445 MWh.year^{-1} for the CE-PST.

The net OpEx cost when using enhanced primary treatment revealed a net saving of £0.13M.year^{-1} for DAF prior to inclusion of capital costs. By contrast, CE-PST was associated with net OpEx of £0.06M.year^{-1}. It is important to note that these results were based on CE-PSTs motivated by phosphorous removal with dosing rates around 30 mg.L^{-1}. If switching to chemical enhancement was purely motivated by a need to deliver load reduction across the primary process to cope with population growth (i.e. increased flow or solid demands), lower quantities of coagulant, estimated at 10 mg.L^{-1} from TWUL's asset standards, will be required providing a net saving of £0.20M.year^{-1}. The CapEx for the DAF plant including dosing plant was estimated at £6.20M for this hypothetical STW serving 500,000 PE which equated to an ACC of £0.26M.year^{-1}. Accordingly, the savings made did not offset the cost of the plant indicating that there is not a convincing case to switch from sedimentation to DAF purely on an economic basis associated with solids and FOG. In comparison, retrofitting chemical dosing to conventional sedimentation processes was found an economically favourable option due to significantly lower capital investment required. However, if the works requires upgrade to flow or load increases that can no longer be resiliently met by the existing sedimentation processes then DAF appears a sensible option. Further, the current economic analysis is based on a low lipid content sludge due to low influent concentrations. Should the FOG levels increase or further optimisation work improve overall solid removal then the case for change can be made of a purely economic basis. For instance, if effluent BOD_5, SS and HEM concentrations respectively as low as 51.3, 73.9 and 10.3 mg.L^{-1}, as obtained with DAF2-FlocB (Table 4), were to be achieved, the current analysis would be adjusted to slightly higher than cost neutrality (Figure 5). In turn, retrofitting conventional sedimentation processes with DAF would be justified from an economic point of view providing significant load reduction to secondary treatment.

The cost associated to installing new PSTs equated to an ACC of £0.99M.year^{-1} whereas retrofitting chemical dosing equated to an ACC of £0.04M.year^{-1} indicating that it provides a feasible economic basis for upgrading primary treatments. Disadvantages of sedimentation tanks include low SORs and hence large footprints and limited ability to control sludge dry solids. In contrast, DAF plants, operated at significantly higher SORs, can be turned up/down by altering the mass of bubbles introduced and can generate thicker sludge with levels appropriate for anaerobic digestion negating the need for thickening processes. These additional features have not been accounted for in the current case but can become critical depending on the specific circumstance of the site in question. In relation to the context of FOG management, upgrading all STWs is not realistic and will require understanding where the benefits would be the highest. Managing FOG at STWs further implies on-going OpEx on sewerage networks. Therefore, more research is required in the field to capture the potential benefits of FOG-control at source to lead to more clarity as to the overall FOG management strategy.

4 Conclusions

Based on a monthly sampling conducted over a five-year period, FOG as HEM was found occurring in urban wastewater at concentrations averaging 57±11 $mg.L^{-1}$. FOG removal efficiencies were reported on average at 50% and 71% respectively from conventional and CE primary sedimentation. By contrast, DAF achieved removal rates of 74% with effluent HEM concentrations of 10±4 $mg.L^{-1}$.

Whilst DAF was evaluated providing significant benefits reducing aeration demand from biological treatment and increasing energy generation through anaerobic digestion, the case to switch from sedimentation to DAF purely on an economic basis was not supported. Yet, DAF, with lower capital investment and footprint required, appeared as a sensible option over sedimentation if the works require upgrading. In relation to FOG management, upgrading all STWs is not realistic. Managing FOG at STWs would imply on-going OpEx in sewerage networks, therefore enhancing primary treatments for FOG removal would require a case-by-case approach to identify where benefits would be the highest.

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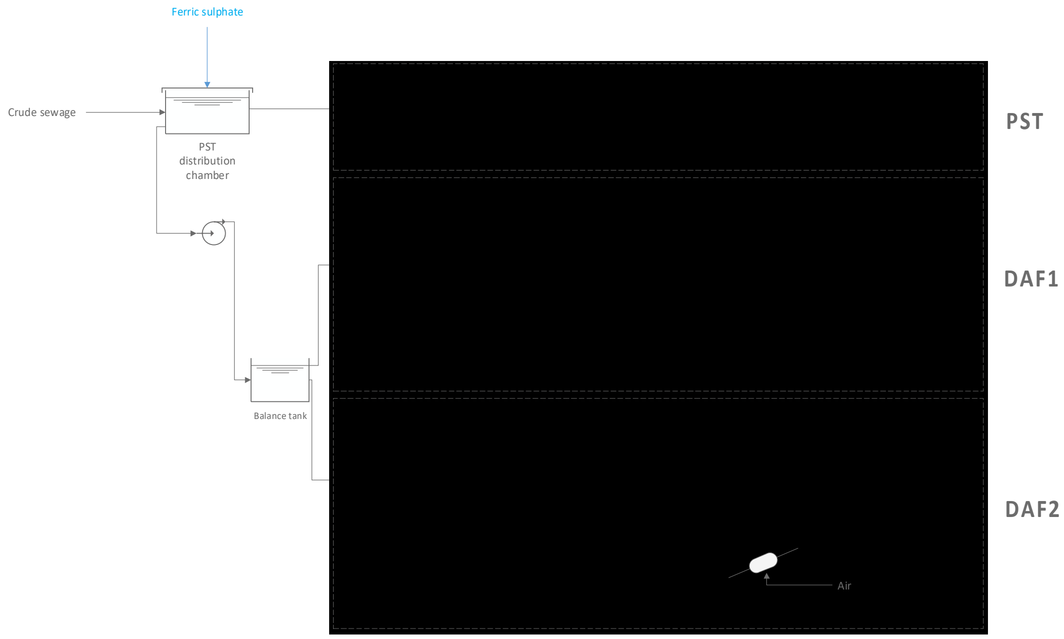

Figure 1 Schematic of the pilot-scale trial.

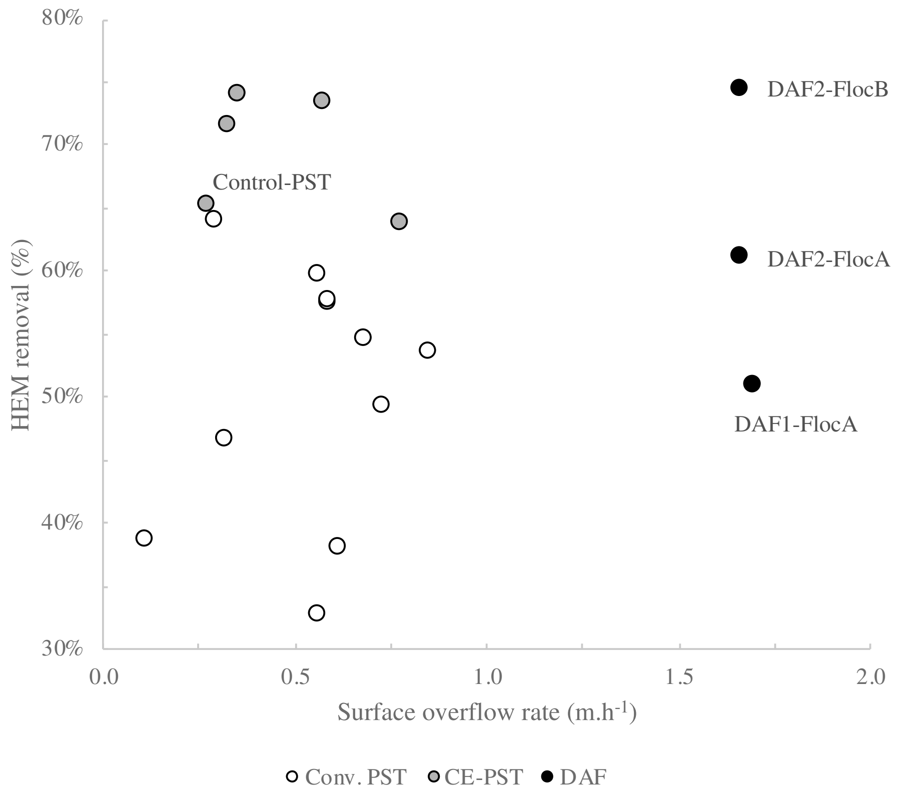

Figure 2 HEM removal rates reported against SOR for each site.

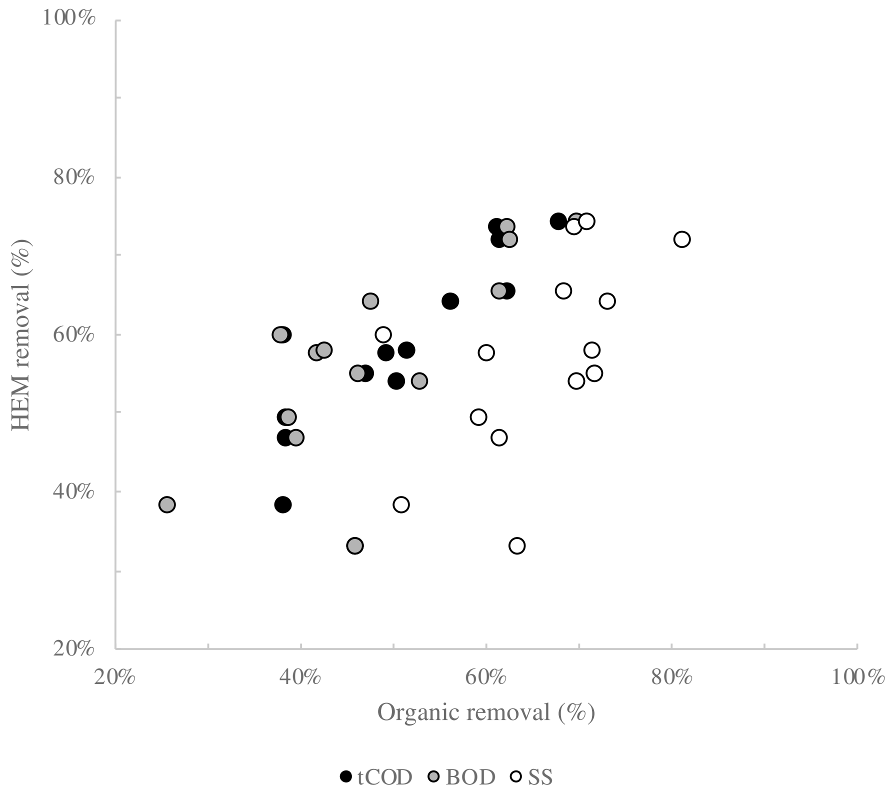

Figure 3 HEM removal rates reported against tCOD, BOD_5 and SS removal rates for each site.

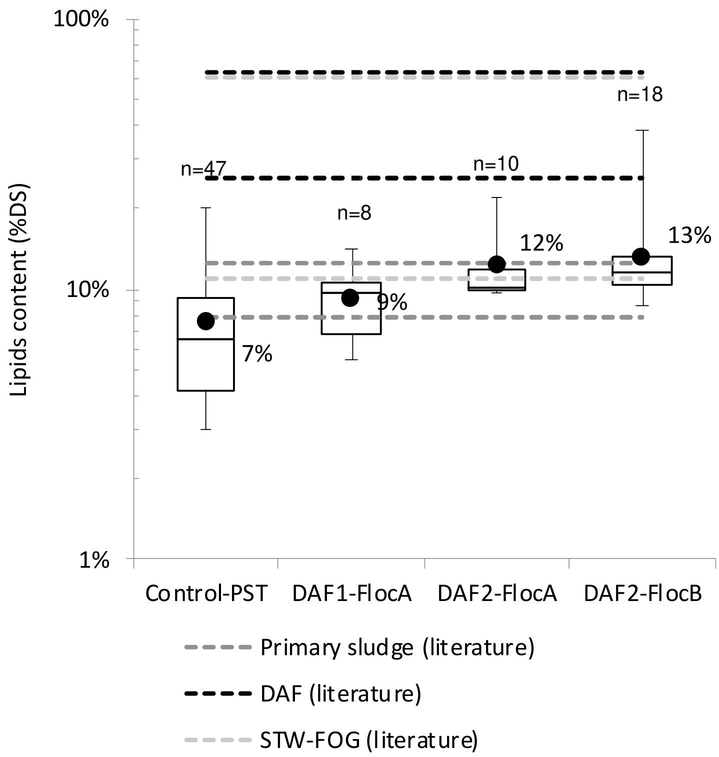

Figure 4 Lipids content (in dry basis) measured in primary sludge from the control-PST and DAF pilot-scale systems.

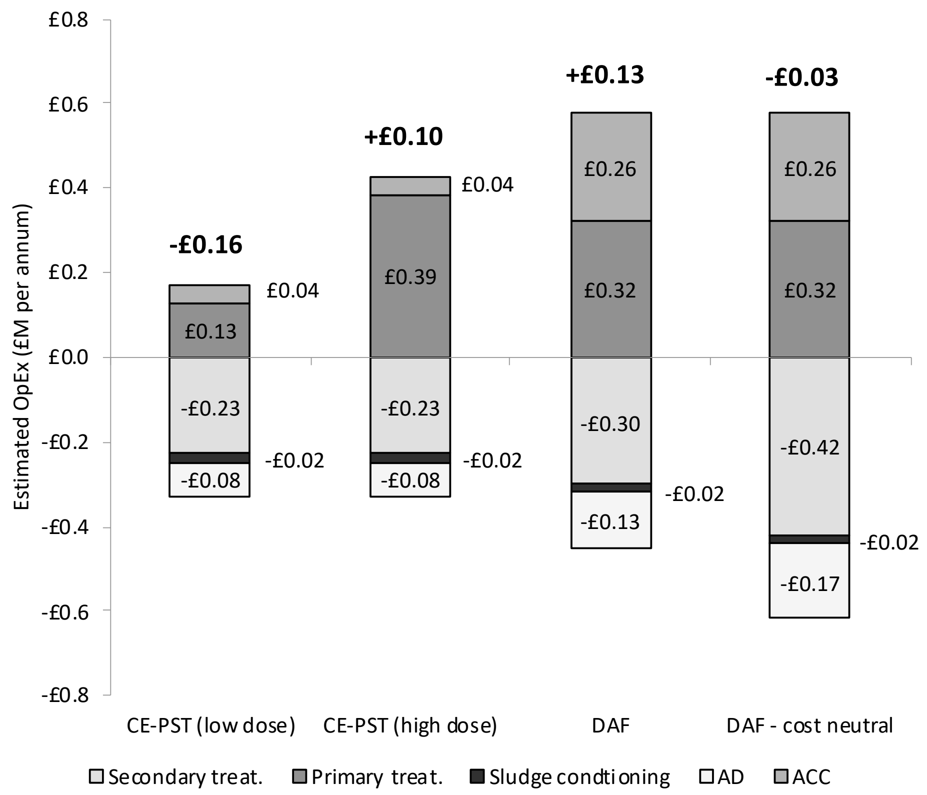

Figure 5 Estimated OpEx savings for CE-PST and DAF from baseline scenario. Positive values indicate costs incurred, whilst negative ones represent savings. Net OpEx are represented in bold for each scenario

Table 1 Key parameters and concentrations of organics for the sampled STWs. Concentrations are expressed as averages with associated standard deviation; tCOD, BOD_5 and SS concentrations were not determined (nd) for sites 14 and 15. Sites 2 to 16 were monitored over a period of five years; site 1 was monitored for four months during the DAF trial. PST type is defined either as conventional (Conv.) or chemically-enhanced (CE).

STW	PE	DWF ($m^3.d^{-1}$)	Solids loading to PST ($kg.d^{-1}$)	SOR ($m.h^{-1}$)	PST type	HEM ($g.capita^{-1}.d^{-1}$)	Concentrations in crude sewage prior treatment ($mg.L^{-1}$)				Concentrations in effluent from primary treatment ($mg.L^{-1}$)			
							HEM	tCOD	BOD_5	SS	HEM	tCOD	BOD_5	SS
1	20,090	3,760	364	0.27	CE	7.5	40±30	452±247	154±84	290±133	14±7	169±49	59±18	91±37
2	123,820	21,970	2,415	0.57	CE	9.3	53±27	574±221	228±82	330±135	12±10	222±96	85±40	100±155
3	130,580	24,570	1,910	0.36	CE	11.9	63±51	724±428	274±233	466±589	16±13	232±87	82±39	135±145
4	156,840	25,590	1,193	0.33	CE	10.3	63±36	658±253	247±83	420±258	18±15	252±85	92±40	78±30
5	927,830	205,740	7,726	0.78	CE	13.4	60±46	578±169	212±57	375±129	22±14	251±61	111±62	100±49
6	89,160	15,240	1,153	0.32	Conv.	10.3	60±29	670±222	265±82	303±94	32±18	411±122	159±66	116±40
7	411,980	100,180	5,221	0.59	Conv.	14.4	59±43	633±240	218±80	417±199	25±15	319±113	127±44	165±177
8	145,410	30,020	4,277	0.62	Conv.	11.9	58±37	670±246	203±72	427±224	36±24	413±237	150±62	209±203
9	180,230	53,600	4,161	0.73	Conv.	11.4	38±37	340±126	130±44	233±340	19±14	208±76	79±25	95±116
10	888,100	192,200	6,119	0.85	Conv.	10.6	49±23	507±139	209±55	255±69	23±13	250±73	98±29	76±21
11	166,770	28,910	1,695	0.57	Conv.	13.4	77±50	663±377	262±97	352±242	31±18	409±154	162±48	179±129
12	221,660	52,910	3,921	0.69	Conv.	18.0	75±42	409±154	287±88	439±192	34±26	398±184	153±77	209±125
13	425,890	88,960	4,052	0.59	Conv.	9.6	46±28	678±186	206±64	364±128	19±12	328±69	117±33	104±32
14	406,400	73,520	3,283	0.30	Conv.	11.5	64±31	774±313	253±84	536±293	23±21	nd	nd	nd
15	227,040	42,640	1,313	0.11	Conv.	10.6	56±24	640±257	251±79	370±291	35±21	nd	nd	nd
16	121,150	28,390	1,614	0.56	Conv.	9.1	39±23	481±186	170±76	284±180	26±21	258±94	92±33	103±54

Table 2 Control-PST and DAF operating parameters.

Parameter	PST	DAF1	DAF2
Flow treated ($m^3.d^{-1}$)	1,221	120	192
Influent solids load ($kg.d^{-1}$)	354	35	56
Screens	N/A	2 mm	N/A
Recirculated water pressure (bar)	N/A	6	3.5
Effective surface area (m^2)	170	2.9	4.8
Recycling ratio (% of inlet flow rate)	N/A	25%	25%
Bubble size (µm)	N/A	20 to 40	10 to 70[1]
Surface overflow rate ($m.h^{-1}$)	0.3	1.7	1.7
Energy consumption ($kWh.m^{-3}$)		0.06	0.07
Air to solids ratio		0.08	0.04

[1] with 90% being between 20 and 50 µm according to the manufacturer

Table 3 List of assumptions used for the economic analysis (primary and secondary treatments).

Parameter	Value	Reference
1 – Primary treatment		
Energy consumption of PST	0.62 $Wh.m^{-3}.d^{-1}$	Newell (2012)
Energy consumption of CE-PST	1.05 $Wh.m^{-3}.d^{-1}$	Newell (2012)
Energy consumption of DAF	70 $Wh.m^{-3}$	DAF2 manufacturer
Coagulant dose for CE-PST	17.3 $g.m^{-3}$	adapted from TWUL asset standards
Cost of ferric sulphate	£344.4 per ton	Kemcore (2019)
2 – Secondary treatment		
BOD of FOG	1.8 kg BOD.kg FOG^{-1}	adapted from Groenewold et al. (1982)
Secondary sludge production	0.8 kg SS.kg BOD^{-1}	TWUL internal data
O_2 demand for BOD_5 removal	0.9 kg O_2.kg BOD_5^{-1}	TWUL internal data
O_2 demand for endogenous respiration	0.04 kg O_2.kg $MLSS^{-1}$	TWUL internal data
Food to microorganisms ratio	0.2	TWUL internal data
Power requirement for aeration	1.5 kWh.kg O_2^{-1}	TWUL internal data
3 – Sludge conditioning		
Polymer dose for thickening/dewatering	10 kg per ton DS	SNF Floerger (n.d.)
Thickening solids capture	95%	Andreoli et al. (2007)
4 – Anaerobic digestion		
Primary sludge destruction	55%	Barber (2014)
Primary sludge biogas yield	0.98 m^3.kgVS $destroyed^{-1}$	Barber (2014)
Secondary sludge destruction	30%	Barber (2014)
Secondary sludge biogas yield	0.79 m^3.kgVS $destroyed^{-1}$	Barber (2014)
COD of FOG	2.8 g COD.g $lipids^{-1}$	Labatut et al. (2011)

COD destruction of FOG	44%	Labatut et al. (2011)
Biomethane yield	0.35 $m^3.kg\ COD^{-1}$	Angelidaki and Sanders (2004)
Calorific value of methane	36 $MJ.m^{-3}$	
Calorific value of biogas	18 $MJ.m^{-3}$	
Electrical conversion efficiency	30%	Goss et al. (2017)
Transportation costs	£8.5 per m^3	ADAS UK Ltd (2013)

Table 4 Average influent and effluents characteristics for HEM, BOD_5, COD and SS with their associated standard deviation. Removal rates were calculated based on average concentrations in influent and effluents. HEM removed and lipids in sludge were calculated and are expressed with their associated uncertainties.

Parameter	Inlet	control-PST	DAF1-FlocA	DAF2-FlocA	DAF2-FlocB
HEM	40±30 n=47	14±7 65% n=22	20±12 51% n=9	16±8 61% n=11	10±4 74% n=17
BOD_5	154±84 n=88	59±18 62% n=83	66±19 57% n=19	67±37 64% n=20	51±13 67% n=26
tCOD	452±247 n=77	169±49 62% n=69	185±54 59% n=19	173±91 62% n=20	158±41 65% n=13
SS	290±133 n=89	91±37 69% n=84	96±27 67% n=19	92±35 68% n=20	74±27 75% n=26
Total P	8.2±3.5 n=66	3.8±0.7 54% n=63	4.1±0.9 50% n=16	3.9±1.1 52% n=17	4.2±1.0 49% n=12

DS (%)	3.1±1.0 n=47	6.6±1.4 n=8	7.1±1.1 n=10	4.9±1.4 n=18
HEM removed (kg.m^{-3} sludge)	3.4±0.4	7.0±1.5	9.3±1.6	6.6±0.7
Lipids in sludge (kg.m^{-3} sludge)	2.3±0.2	5.7±0.8	8.8±1.1	6.7±0.9

Table 5 Energy required for aeration and generated through anaerobic digestion. The base case considers a conventional PST. Positive values indicate savings while negative ones represent demands.

Parameter (in MWh.year^{-1})	PST	CE-PST	DAF
Energy demand from primary treatment	-23	-38	-2,555
Total energy demand for BOD_5	-7,763	-5,220	-4,417
Energy demand for FOG	-2,526	-1,465	-665
Total energy from anaerobic digestion	+6,137	+7,054	+7,592
Energy generation from anaerobic digestion of FOG		+375	+651
Net energy	-1,649	+1,796	+620
Net change from base case		+3,445	+2,269

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